PB84-199868

Application of the Dual Alkali Process at a 280 MW Coal-Fired Power Plant

(U.S.) Industrial Environmental Research Lab.
Research Triangle Park, NC

Jun 84

U.S. Department of Commerce
National Technical Information Service
NTIS

PB84-199868

EPA-600/D-84-145
June 1984

APPLICATION OF THE DUAL ALKALI PROCESS AT A 280 MW COAL-FIRED POWER PLANT

by

Norman Kaplan
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina 27711

INDUSTRIAL ENVIRONMENTAL RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OH 45268

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

TECHNICAL REPORT DATA *(Please read Instructions on the reverse before completing)*		
1. REPORT NO. EPA-600/D-84-145	2.	3. RECIPIENT'S ACCESSION NO. PB84 199868
4. TITLE AND SUBTITLE Application of the Dual Alkali Process at a 280 MW Coal-fired Power Plant		5. REPORT DATE June 1984 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Norman Kaplan		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS See Block 12		10. PROGRAM ELEMENT NO. 11. CONTRACT/GRANT NO.
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Published paper· 3/79 - 5/81 14. SPONSORING AGENCY CODE EPA/600/13
15. SUPPLEMENTARY NOTES Author Kaplan's Mail Drop is 61; his phone is 919/541-2556.		

16. ABSTRACT The paper gives results of applying the dual alkali (D/A) flue gas resulfurization (FGD) process to a 280-MW coal-fired power plant. (NOTE: D/A is a generic term applied to FGD systems that use soluble alkali to absorb SO_2 and then react the spent scrubber solution with lime and/or limestone to regenerate the alkali and produce a waste calcium-sulfur salt for disposal. Features that distinguish D/A from conventional direct lime and limestone scrubbing include: (1) use of a clear solution, rather than a slurry, for contacting the flue gas in the absorber; and (2) reaction of the solution in a separate absorbent regeneration section to form the waste solids, rather than forming the waste solids as a part of the scrubbing operation.) The FGD system met all acceptance test process guarantees except filter cake properties (averaging 52.2 of a minimum required 55 wt % insoluble solids). The remaining guarantees were: SO_2 emissions limit, particulate matter emissions, lime consumption, soda ash consumption, power consumption, and system availability. The demonstration started on May 1, 1980, and ended on May 31, 1981.

17. KEY WORDS AND DOCUMENT ANALYSIS			
a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution	Coal	Pollution Control	13B 21D
Flue Gases	Calcium Oxides	Stationary Sources	21B 07B
Desulfurization	Sodium Carbonates	Dual Alkali Process	07A, 07D
Alkalies			
Scrubbers			13I
Electric Power Plants			10B
18. DISTRIBUTION STATEMENT Release to Public		19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 35
		20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (9-73)

NOTICE

This document has been reviewed in accordance with U.S. Environmental Protection Agency policy and approved for publication. Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

INTRODUCTION

Since 1972, the United States Environmental Protection Agency (EPA) has been involved in the development of dual (or double) alkali (D/A) flue gas desulfurization (FGD) technology as an improved alternative to wet lime and limestone slurry scrubbing technology which was the most prevalent technology then in use for coal-fired utility boilers. Expected improvements with D/A were higher SO_2 removal, with lower liquid-to-gas ratio (L/G), and reduction in scaling and plugging problems in FGD equipment caused by the use of slurries containing insoluble calcium compounds. The EPA has carried on an extensive development program involving engineering and economic studies, and testing at successively larger scales as confidence in the technology grew. Testing was conducted at laboratory, pilot plant, industrial boiler, and utility prototype (20 MW) prior to the embarkation into a full-scale utility (280 MW) test program which is the subject of this paper, and the final stage of development of this technology.

The full-scale utility demonstration program was a joint effort by several organizations under the sponsorship of the EPA. The system was designed by the air pollution control group of Combustion Equipment Associates (CEA)[a] and Arthur D. Little, Inc. (ADL) and installed on Unit 6 of Louisville Gas and Electric Company's (LG&E) Cane Run Station in Louisville, Kentucky. Unit 6 was a coal-fired boiler burning high sulfur coal without any means of SO_2 control. Installation was by the construction department of LG&E, with technical guidance on process design from CEA.

The D/A system was designed to control SO_2 emissions to less than 200 ppm (dry basis) without additional air dilution with the boiler burning coal containing up to 5% sulfur. When coal containing greater than 5% sulfur is fired, the system is designed to remove at least 95% of the SO_2 in the inlet flue gas. The D/A system is not designed for removal of particulate matter. It is designed, however, not to increase the loading of particulate matter in the flue gas.

As a demonstration system, the purpose of the installation and operation was to establish:

- Overall performance - with respect to SO_2 removal, lime consumption, soda ash consumption, regeneration of spent liquor, water balance, scaling or solids buildup, materials of construction, waste cake properties, and system dependability.

- Economics - capital investment and operating cost.

[a]Now part of Thyssen-CEA Environmental Systems, Inc.

Performance Guarantees

The process designer offered LG&E seven performance guarantees. These performance guarantees, summarized in Table 1, apply to SO_2 emissions, particulate matter emissions, lime (calcium) consumption, soda ash consumption, power consumption, waste solids properties, and system availability.

Construction of the FGD system was completed in March 1979, and the system was initially charged and started up in April 1979. Various problems with the boiler, the FGD system, and the quality and quantity of lime supplied to the system delayed the acceptance testing until after the demonstration period had begun. The acceptance test was successfully conducted in July 1980, when Unit 6 and all FGD system components were fully and simultaneously available, and the boiler could be operated at (or near) full load. The demonstration period officially started on May 1, 1980, and ended on May 31, 1981.

The D/A FGD process was developed to overcome the disadvantages of lime and limestone scrubbing (such as scaling) while retaining the cost advantages of a throwaway system. The use of a scrubbing solution instead of a slurry allows D/A FGD to: (1) achieve high SO_2 removal efficiencies over a wide range of inlet SO_2 concentrations (with alkaline solutions); and (2) minimize scale potential in the scrubber circuit. Precipitation of the waste calcium sulfite/sulfate solids outside the scrubber facilitates the formation of waste solids with good dewatering properties. In addition, it is not necessary to wash the mist eliminator to prevent solids deposition and scale formation.

GENERAL DESCRIPTION AND PROCESS CHEMISTRY

D/A is a generic term applied to FGD systems that use soluble alkali to absorb SO_2 and then react the spent scrubber solution with lime and/or limestone to regenerate the alkali and produce a waste calcium-sulfur salt for disposal.

The two principal features of D/A technology that distinguish it from conventional direct lime and limestone scrubbing are: (1) the use of a clear solution rather than a slurry for contacting the flue gas in the absorber; and (2) reaction of the solution in a separate absorbent regeneration section to form the waste solids rather than forming the waste solids as a part of the scrubbing operation.

Dilute vs. Concentrated Mode of Operation

D/A FGD systems are usually classified as dilute or concentrated, depending on the concentration of active alkalis in the scrubbing liquor. Active alkalis are ions in the scrubbing liquor that participate in the SO_2 absorption reactions, and they include $CO_3^{=}$, HCO_3^{-}, OH^{-}, $SO_3^{=}$, and HSO_3^{-}. (Although it does not react directly with SO_2, HSO_3^{-} is considered an active alkali because it can be converted to an active form by reaction with lime or limestone.)

TABLE 1. PERFORMANCE GUARANTEES

Parameter	Guarantee
SO_2 Emissions Limit	200 ppm (dry basis) without additional air dilution (for up to 5.0% sulfur coal)
Calcium Consumption	1.05 moles available CaO per mole SO_2 removed
Soda Ash Consumption	0.045 moles Na_2CO_3 per mole SO_2 removed
Particulate Matter Addition	No net particulate matter addition by FGD system
Power Consumption	System to consume not more than 1.2% of power generated at peak capacity (excluding reheat)
Filter Cake Properties	Filter cake to contain a minimum of 55% (weight basis) insoluble solids
System Availability[a]	The system shall have an availability (as defined by the Edison Electric Institute) of at least 90% for 1 year. (The Edison Electric Institute defines operating availability as "the percent of time the unit was available for service, whether operated or not. It is equal to available hours divided by the total hours in the period under consideration and is expressed as a percentage.")

[a]This guarantee was amended in May 1980 in a contract between EPA and LG&E to require an operability of 90% or greater, coupled with a daily average SO_2 removal efficiency of 90% or greater.

The concentration of Na^+ (sodium ion) in solution that balances the active alkali is known as active sodium (although the Na^+ itself does not participate in the reactions). Dilute systems generally have active sodium concentrations less than 0.15 molar; in concentrated systems the concentrations are greater than 0.15 molar. This distinction is important when determining appropriate methods for sulfate removal. Concentrated systems operate unsaturated with respect to gypsum ($CaSO_4 \cdot 2H_2O$), and gypsum does not precipitate as a separate crystalline phase during regeneration. In concentrated systems, sulfate ions coprecipitate with calcium sulfite during regeneration as a mixed crystal of calcium sulfate and calcium sulfite. In dilute systems, on the other hand, sulfate precipitates during regeneration as gypsum.

Scrubber oxidation rate is an important factor in determining the use of concentrated vs. dilute D/A systems. Concentrated systems are generally best suited to applications with low oxidation rates -- typically those where medium- to high-sulfur coals are burned with little excess air. Dilute systems are preferable where oxidation rates are high. Burning low-sulfur coal promotes high oxidation rates because of the high O_2/SO_2 ratio in the flue gas. Boiler firing with excess air has a similar effect. The D/A installation at LG&E was designed to operate in the concentrated mode because of the high-sulfur coal used at Cane Run.

Flue Gas Scrubbing

In the flue gas scrubbing section, SO_2 is removed as the flue gas is contacted with the alkaline scrubbing solution on the trays of the absorber. The major active constituent of the scrubbing liquor, sodium sulfite, reacts with SO_2 in the flue gas to form bisulfite:

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (1)$$

Small amounts of hydroxide and carbonate may also be present in the regenerated scrubbing liquor from the thickener. To reduce the potential for calcium carbonate scale formation on the trays, a portion of the slightly acidic liquor from the bottom of the absorber is blended with the alkaline regenerated liquor from the thickener to maintain a pH of below 10. This slightly acidic recirculated liquor reacts with the carbonate and hydroxide to yield additional sulfite:

$$Na_2CO_3 + 2NaHSO_3 \rightarrow 2Na_2SO_3 + H_2O + CO_2\uparrow \qquad (2)$$

$$NaOH + NaHSO_3 \rightarrow Na_2SO_3 + H_2O \qquad (3)$$

Oxidation of the sodium sulfite to sodium sulfate:

$$2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4 \qquad (4)$$

must be held to less than 20% by limiting: (1) oxygen concentration in the flue gas or (2) the number of absorber modules operated at low load. Otherwise, the rate at which sulfate can be purged as a mixed crystal will be exceeded, and the solids settling rate will deteriorate as the system approaches dilute mode conditions.

Absorbent Regeneration

The active sodium in the absorber is defined as the sodium ion concentration that balances the active ($SO_3^=$, OH^-, $CO_3^=$, HCO_3^-) or spent (HSO_3^-) anionic species. Sulfate ($SO_4^=$) is inactive for SO_2 absorption.

$$[\text{active sodium}] = 2 \times [Na_2SO_3] + [NaOH] + 2 \times [Na_2CO_3] + [NaHSO_3] + [NaHCO_3]$$

The activity of the spent liquor from the absorber is restored by reaction with lime slurry in a two-step process:

$$Ca(OH)_2 + 2NaHSO_3 \rightarrow CaSO_3 \cdot 1/2H_2O + Na_2SO_3 + 3/2H_2O \quad (5)$$

$$Ca(OH)_2 + Na_2SO_3 + 1/2H_2O \rightleftarrows CaSO_3 \cdot 1/2H_2O + 2NaOH \quad (6)$$

Equation 5 represents a neutralization reaction that goes to completion. The equilibrium hydroxide concentration shown in Equation 6 is limited by the relative solubility of calcium sulfite and calcium hydroxide and the concentrations of hydroxide. The reactions shown in Equations 5 and 6 lead to precipitation of calcium sulfite hemihydrate.

A limited amount of calcium sulfate is also precipitated as a mixed crystal of calcium sulfite/sulfate:

$$xSO_3^= + ySO_4^= + (x+y)Ca^{++} + zH_2O \rightarrow Ca_{(x+y)} \cdot (SO_3)_x \cdot (SO_4)_y \cdot zH_2O \quad (7)$$

where the ratio of x to y is usually greater than 4:1.

In a concentrated mode, the sulfate coprecipitates with the calcium sulfite, and gypsum (the dihydrate) is not formed. This allows the system to operate unsaturated with respect to calcium sulfate and reduces the potential for gypsum scaling.

The regeneration at Cane Run Unit 6 is accomplished in a two-stage reactor system that prolongs the reactor residence time and produces solids with improved settling and filtering characteristics.

Solids Dewatering

Solids dewatering involves thickening the reactor effluent slurry to 15 to 30-wt % solids, followed by filtration to produce a moist waste filter cake. While the filter cake is being formed, it is washed with fresh water to recover sodium salts and prevent sodium loss in the process liquor discharged with the cake. The wet filter cake constitutes the only waste discharged from the D/A process.

The solids content and chemical composition of the waste cake depend primarily on the amount of oxidation, the quality of lime, the operation of the filter, and the chloride content of the coal. For utility boilers firing medium- to high-sulfur coal, the waste cake for a system operating with commercial lime is expected to contain 55 to 70-wt % insoluble solids and have a chemical composition with the following ranges:

Species	Weight % (Dry Basis)
$CaSO_3$	80 to 85
$CaSO_4$	10 to 15
$CaCO_3$ + Inerts	5 to 10
Soluble Sodium Salts	1 to 2

The amount of sodium lost in the cake depends on the sodium concentration in the process liquor, the amount of oxidation, the chloride content of the coal, the operation of the filter, and the permissible level of cake washing.

DESCRIPTION OF THE D/A SYSTEM AT LG&E CANE RUN NO. 6

General

Cane Run Unit 6 consists of a pulverized coal-fired steam boiler (built by Combustion Engineering), with a Westinghouse turbine-generator. The unit was designed to operate from a minimum of 60 MW during off-peak hours to a maximum of 300 MW during peak hours. The annual average load is equivalent to approximately 180 MW (about 60% of the gross peak capacity).

Flue gas from the boiler passes in parallel streams through two Ljungstrom combustion air preheaters. Each preheater discharges flue gas through separate ducts to an electrostatic precipitator designed for 99.4% removal efficiency of particulate matter. From the precipitator, the gases enter two parallel induced draft fans, each handling 50% of the total gas.

The SO_2 removal system, installed between the existing induced draft fans and the stack, takes hot flue gas from the outlet of the boiler induced draft fans. Two forced draft booster fans push the flue gas from the ID fans through the absorber. The scrubbed gas from the absorbers is reheated with hot air from an oil-fired burner and then returned to the existing entrance to the stack. In case of scrubber malfunction, dampers have been provided to allow gas bypass around the scrubber system via the existing ductwork.

Coal for Unit 6 is received from several sources. The average sulfur content on a dry basis is approximately 3.5%. (The design for the LG&E dual alkali system is based on 5% sulfur coal.) The average chloride content of the coal is 0.04%, ranging from 0.03% to 0.06%.

D/A FGD System

The LG&E D/A system consists of five process sections:

- o Flue gas scrubbing.
- o Absorbent regeneration.
- o Solids dewatering.
- o Raw materials preparation.
- o Waste disposal.

Process flow for flue gas scrubbing is shown in Figure 1; absorbent regeneration, solids dewatering, and raw materials preparation in Figure 2; and waste disposal in Figure 3.

Flue Gas Scrubbing - The flue gas scrubbing section (Figure 1) consists of two identical scrubber modules, each containing a booster fan, an absorber, an oil-fired flue gas reheater, and two recirculation pumps (one operating and one backup).

Flue gas is drawn from the existing induced draft fans and is forced through the 2-stage tray absorbers by the booster fans. Although both absorbers were designed to operate simultaneously (with each handling half the boiler load), one absorber is capable of handling all of the gas during periods of low boiler load. To allow for greater flexibility in operation and provide for the possibility of maintenance on one module while the system is in use, a common duct connecting the two booster fan inlets was incorporated. A bypass was also provided to allow complete shutdown of the scrubbers while the boiler is still on-line. The bypass and FGD system inlet dampers were interlocked to enable the absorbers to be brought on- or off-line without interrupting boiler operation.

Hot flue gas (149°C) entering each absorber is first cooled by a set of sprays that direct scrubbing solution to the underside of the bottom tray. In addition to providing temperature control at the bottom of the absorbers, these sprays minimize accumulation of fly ash solids on the underside of the tray and the bottom of the absorber. The cooled, saturated gas passes through a set of two trays where SO_2 is removed by absorbent liquor that circulates countercurrent to the gas. The flue gas is then routed through a chevron-type mist eliminator to remove any entrained absorber liquor droplets. After leaving the absorber, the scrubbed gas is mixed with hot air from an oil-fired reheater to produce 28°C of reheat (to a temperature of about 79°C) to prevent condensation and corrosion.

After the scrubbing solution passes the internal contacting area (two single-pass crossflow perforated plates), it is collected at the bottom of the absorber. Part of the liquor is recirculated to the sprays and part to the trays. Recirculating liquor and regenerated liquor are flow-mixed prior to entering the top tray. The flow rate of the recirculating liquor is adjusted to keep the tray pH below 10. This minimizes the absorption of CO_2 and decreases the potential for $CaCO_3$ scaling in the absorber.

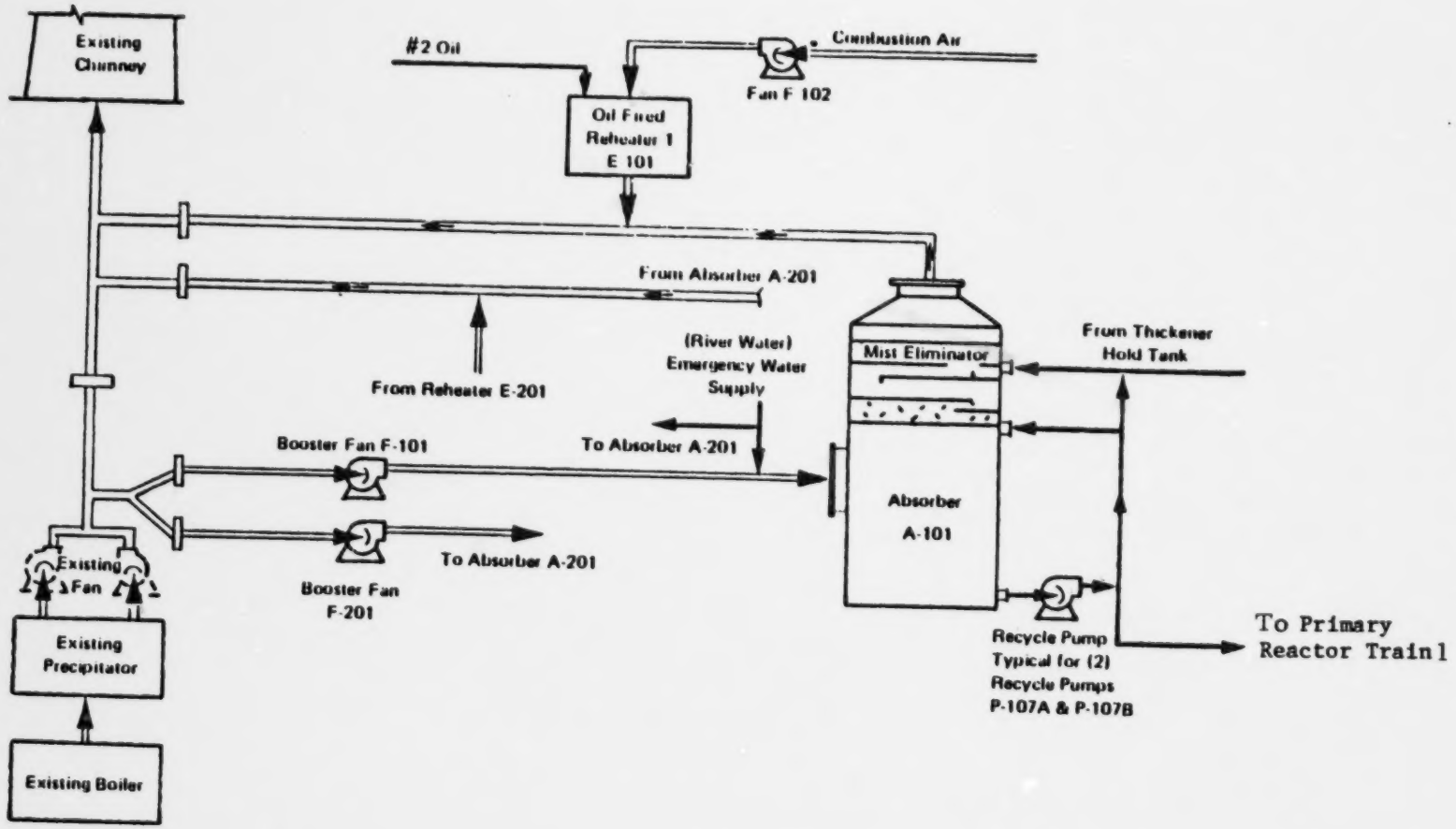

FIGURE 1. LG&E DUAL ALKALI FGD SYSTEM: FLUE GAS SCRUBBING SECTION

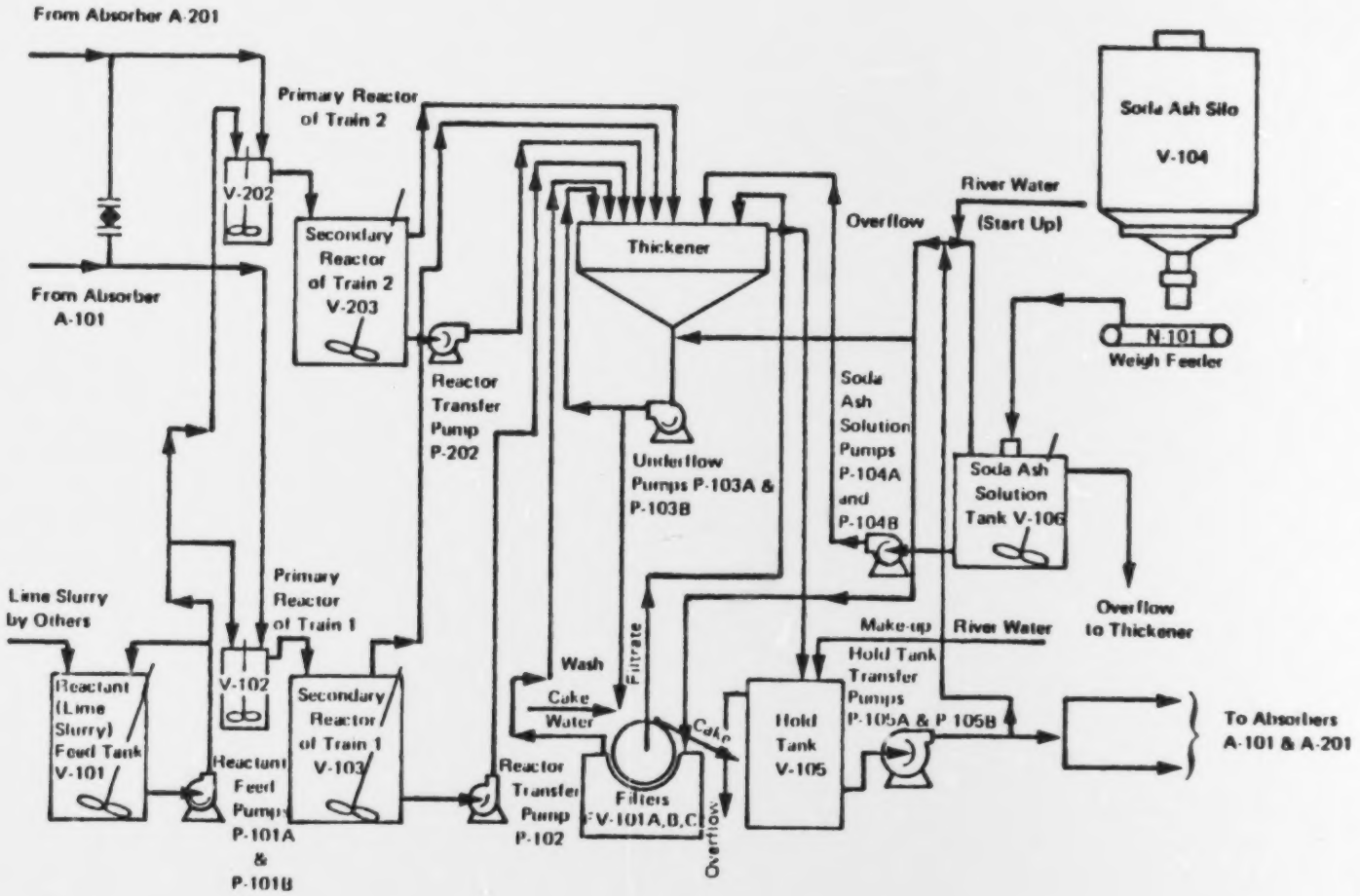

FIGURE 2. LG&E DUAL ALKALI FGD SYSTEM: ABSORBENT REGENERATION SECTION

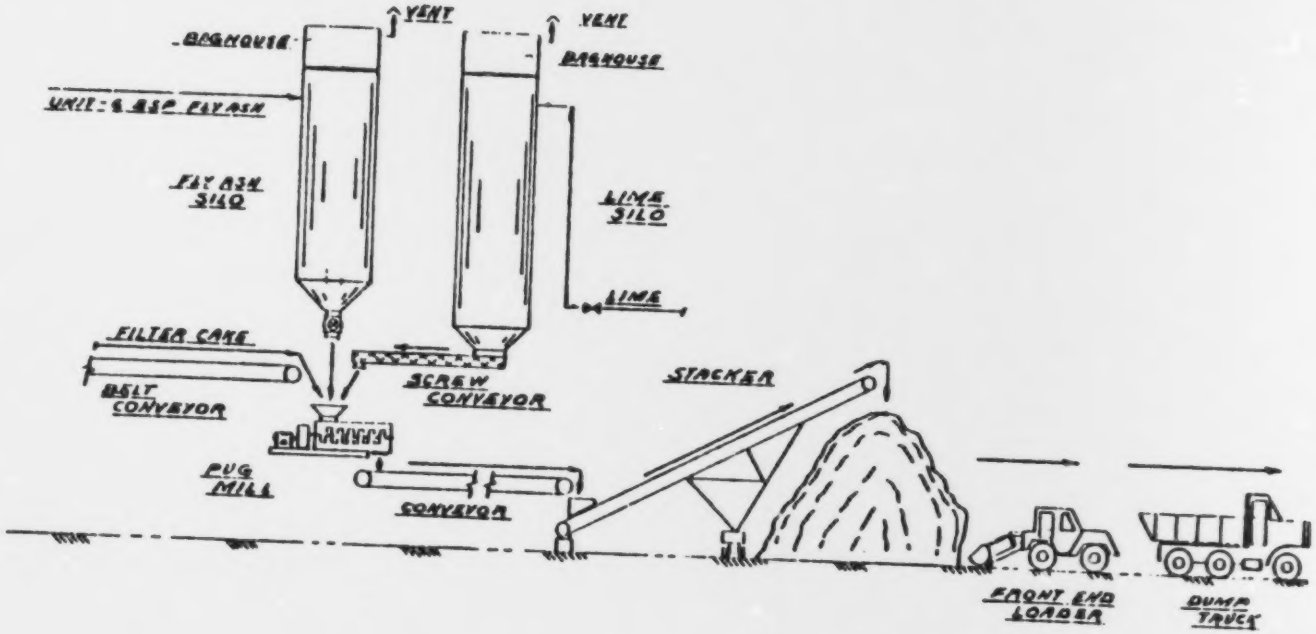

FIGURE 3. LG&E DUAL ALKALI SLUDGE TREATMENT SYSTEM

An automatic emergency water supply is provided to quench the hot flue gas and prevent damage to the absorber lining in the event of recycle pump failure. Pressures and temperatures are measured at appropriate points throughout the system. SO_2 removal is monitored by continuous SO_2 and O_2 analyzers at the inlet of the absorbers and in the stack. The pH's of the spent liquor from the absorber and of the top tray feed are also continuously monitored.

Absorbent Regeneration - This section of the system (Figure 2) consists of two parallel reactor trains. Each contains a primary and secondary reactor. When operating under design conditions (i.e., full load), each reactor train serves to regenerate solution from one absorber. For short periods of time, a single reactor train can handle the solutions from both absorbers operating at full load. When the boiler is firing typical or average coal (3.5% sulfur), only one reactor train is normally required.

The spent scrubbing solution is introduced to the primary reactor along with slurried carbide lime from the lime day-tank. In this reactor (3 to 15 minutes residence time), the regeneration of scrubbing solution begins. The primary reactor overflows into a larger secondary reactor (30 to 60 minutes residence time) where regeneration and crystal formation are completed. The secondary reactor was designed for a pH range of 11 to 12 maintained by controlling the amount of lime slurry fed to the primary reactor. In later operation, problems due to alumina and silica impurities in the carbide lime were alleviated by operating at a pH of 9.0 in the secondary reactors.

Solids Dewatering - The slurry from the secondary reactor (Figure 2) is fed to the thickener feed well to initiate separation of the waste solids from the regenerated solution. The thickener is operated to provide 15 to 30-wt % solids in the underflow, even though the slurry can settle to 40-wt % solids or more. Slurries of about 25-wt % solids allow better control of filter cake washing on rotary drum vacuum filters, control of which cannot always be achieved with variations in filter drum speed and vat depth alone. The thickener underflow slurry is recirculated past the filters in a recycle loop that feeds a portion to the filters and returns the remainder to the thickener feed well. Each filter is equipped with an overflow pipe that returns slurry to the thickener to permit operation of the filter in an overflow mode.

There are three filters at LG&E, each rated to handle 50% of the total solids produced at the design conditions. Each filter can be operated independently. For optimum performance (i.e., to obtain cake containing high percentage dry solids and low percentage soluble salts), the filters should run at fixed conditions (constant drum speed, submergence, wash ratio, etc.). Therefore, the cake discharge rate is controlled by changing the number of filters in operation. The number of filters to be placed on line is determined by the quantity of solids accumulated in the thickener. This in turn is judged from the solids concentration in the underflow slurry and the thickener rake lift position indicator. The density of the underflow slurry is measured by a densitometer. (The

density of the underflow is monitored by a gamma radiation emission source/detector instrument that measures radiation absorption of the slurry in the pipeline and translates this signal into specific gravity.) Thickener (overflow) hold tank liquor is added at the thickener underflow outlet as required to maintain the solids concentration in the underflow slurry at about 20 to 25%. The number of filters in use is changed if the concentration of solids in the underflow slurry cannot be controlled with the dilution liquor.

The waste cake is washed on the filter by a series of water spray banks. This wash is designed to remove a large fraction (approximately 90%) of the occluded soluble salts from the cake and to return these salts to the system, thereby reducing sodium losses and minimizing sodium carbonate makeup. The total wash rate is usually set at a constant percentage of the cake discharge rate. The mixed filtrate and wash liquor from the filter are collected in the filtrate sump and returned to the thickener. Overall, the total allowed wash rate is controlled by the water balance on the rest of the FGD system.

Clear liquor overflow from the thickener is collected in the thickener hold tank, which provides surge capacity for the absorbent liquor feed to the scrubber system and maintains overall control of the volume of liquor in the system. Water is added to this hold tank to make up for the difference between total system water losses (evaporation and cake moisture) and total water inputs from other sources (pump seals, lime feed, and cake wash).

Raw Materials Preparation - Two chemicals are required to operate the D/A system at Cane Run Unit 6: carbide lime for liquor regeneration and soda ash to compensate for sodium losses. The lime slurry tank and soda ash feed system are shown in Figure 2.

Carbide lime, a byproduct of acetylene production, is barged to the Cane Run Station as a slurry containing approximately 25% insoluble solids. The slurry is pumped to a main storage tank at the plant. Excess liquor is then decanted from the tank and reused to transport the carbide lime. From the main storage tank, the lime slurry is pumped to a classifying and grinding system consisting of a hydroclone and wet ball mill. The fine fraction of lime is then routed to the D/A system day-tank, which supplies lime to the primary reactor at the appropriate rate. The coarse fraction is ground in the wet ball mill and sent to lime storage for use in the lime scrubbers on Cane Run Units 4 and 5.

The system was designed to utilize 95% of the available $Ca(OH)_2$ in the carbide lime slurry. Thus the overall feed stoichiometry (moles available $Ca(OH)_2$ per mole of SO_2 absorbed) is expected to be 1.05. The carbide lime as delivered to the D/A system contains about 92.50% $Ca(OH)_2$ and 7.5% inerts on a dry basis.

Dry, dense soda ash is received at the plant and stored in the soda ash silo, which feeds the soda ash solution tank by means of a weigh feeder (Figure 2). The rate of addition is proportional to the amount of lime slurry fed to the system. This relationship exists because the sodium losses in the cake can be expressed as a percentage of the solids produced, which in turn is related to the amount of lime added to the system. The overall rate of soda ash additions is also periodically adjusted based on analysis of the absorbent liquor.

Soda ash liquor can be made up with either fresh water or clarified liquor drawn from the thickener hold tank. Provisions have been made to feed the soda ash solution either directly to the absorbers or to the thickener. The normal mode of operation is to prepare the soda ash makeup solution using clarified (thickener overflow) liquor and to feed it to the absorbers.

Waste Disposal - A long-range disposal plan for the D/A filter cake was developed as part of an overall disposal program for all FGD wastes produced at Cane Run. The plan involves stabilization of the wastes generated by each of the three FGD systems via the addition of lime and fly ash, followed by dry landfill of the stabilized material.

The Conversion Systems, Inc. (CSI) waste processing plant is installed at the end of the filter building (Figure 3). Filter cake from the dual alkali system discharges into a single conveyor (the three filters are arranged in a single row, end-to-end), which carries the cake directly to the pug mill mixer in the processing plant. A belt weigh element near the end of the conveyor measures the quantity of cake produced. Fly ash from Unit 6 is pneumatically transported to the processing plant and stored in a live-bottom silo above the pug mill. From there it is discharged through a star valve into the pug mill. The fly ash feed is adjusted according to the filter cake discharge rate and amounts to approximately 60% by weight of the dry filter cake.

Quicklime is also fed to the pug mill at a rate equivalent to 2 to 3% of the dry weight of the cake. The quicklime is stored in a dry silo and fed to the pug mill by screw conveyor.

The pug mill discharges to a radial stacking conveyor. The stacker spreads the material in a windrow (interim stockpiling area) where it remains for 1 to 3 days to initiate the stabilization reactions. The material is then removed from the windrow and loaded into rear dump trucks with a front end loader. The trucks transport the material to a dry landfill. Provisions have also been made for direct loading of the trucks from the stacking conveyor.

PROCESS DESIGN AND WATER BALANCE

The D/A system at LG&E was designed to accommodate the characteristics and volume of flue gas to be treated from Cane Run 6 with the types and sources of raw materials available at this location.

Design Basis

The design basis for the D/A system at LG&E is summarized in Table 2. Additional details of the design basis were presented in the Phase I report for this project (EPA-600/7-78-010).

Water Balance

The D/A system was designed to operate in a closed-loop mode, with no solid or liquid discharge other than that present in the filter cake waste product.

The thickener hold tank was designed to maintain overall control of the volume of liquor in the system. The hold tank collects clear liquor overflow from the thickener and provides surge capacity for the absorbent liquor feed to the scrubber system. Water is added to the thickener hold tank to compensate for the difference between total system water losses (evaporation and filter cake moisture) and total water inputs from other sources (sodium makeup solution, pump seals, lime feed, and cake wash).

Table 3 shows the overall water balance at design conditions. Process water enters the system in the lime slurry (both as slurry water and chemically combined water), cake wash, instrument purge, and pump seals. (All the water used for pump seals eventually enters the system. The water that drains out of the seals is collected in sumps and returned to the thickener.)

More than 75% of the water lost from the system at design conditions is due to evaporation in the flue gas. The extent of evaporation in the absorbers depends on the boiler load and the temperature and humidity of the inlet flue gas. Water is also removed from the system with the waste cake, both as free water and as chemically bound water.

In addition, water enters and leaves the system via liquid surfaces exposed to the atmosphere (mainly the thickener and thickener hold tank). Based on meteorological conditions at Louisville, thickener and hold tank configurations, and average liquor temperatures, a significant net evaporation of water from the system was anticipated. The annual average evaporation rate from the thickener and hold tank was estimated at 121 liters/min. and the average gain from rainfall at about 2 liters/min.

Since water leaving the system would on the average exceed the uses of water, a net amount of makeup was considered necessary to maintain the system water balance. This was to be added to the thickener hold tank as required to maintain a constant liquid level. The average rate of makeup addition was estimated at 440 liters/min. for the following design conditions: boiler operating at full capacity (300 MW) and firing 5% sulfur coal.

TABLE 2. DESIGN BASIS FOR THE D/A SYSTEM AT LG&E'S CANE RUN UNIT 6

Boiler Conditions	
Boiler Gross Peak Capacity, MW	300
Boiler Gross Net Capacity, MW	277
Annual Load Factor, %	60 (of gross peak capacity)
Basis-input heat/gross peak electricity generated (kJ/kWh)	10,500
S in Coal, %	5.0
Cl in Coal, %	0.04
Ash, %	17
Moisture, %	9
Coal Energy Content (J/g)	26,000
Gas Flow Rate @ Scrubber Inlet (m^3/min)	30,200
Temperature, °C	150
SO_2, kg/min.	177
Fly Ash, kg/min.	1.8
SO_2 @ Scrubber Inlet (ng/J)	3,680
SO_2 Removal System	Retrofit
SO_2 @ Scrubber Outlet (ng/J)	215
SO_2 Removal in Scrubber, %	94.2
Particulate Removal	None
Bypass	Yes
Turndown	to 20% of Capacity
Number of Absorber/Reactor Modules	2
Number of Filter Units	3
Thickeners	1
Carbide Lime Feed System	Existing (new day tank)
Soda Ash Silo Capacity	14 days
Reheat	Direct Reheat with Oil-Fired Reheater
Sludge Disposal	On-Site Pond with Treatment[b]
Spare Capacity (Percent)[a]	
Filters	50
Pumps	100
Necessary Instruments to Run Process	100

[a] The system can be operated with one reactor train, if necessary, at full load and one absorber system at less than 60% load.

[b] Later changed to CSI sludge treatment and landfill.

Source: EPA-600/7-78-010.

TABLE 3. OVERALL WATER BALANCE AT DESIGN CONDITIONS

Process Water	Inlet (liters/min.)	Outlet (liters/min.)
Lime Slurry		
- Free Water	488	
- Chemically Combined	47	
Filter Cake Wash	343	
Pump Seals	76	
Instrument Purge	2	
Rain	2	
Makeup (to thickener hold tank)	440	
Evaporation		
- Absorber		1053
- Open Tanks		121
Filter Cake		
- Free Water		201
- Chemically Combined		23
	1398	1398
Non-Contact Water		
Fan Bearing Cooling	662	662
Vacuum Pump Seals	151	151
	813	813

The process design included the following specific provisions to avoid inadvertent discharge of process water:

- Drains are provided at all pump stations and inside the filter building to direct all process pump seal water leaks, pump and piping flush water, and equipment and building washdown water to the filtrate sump for return to the thickener along with the filtrate and cake wash water.

- A separate sump is provided to collect vacuum pump seal water. This was intended to be non-contact water, but provisions have been made to pump it to the thickener if it becomes contaminated with process liquor.

- The emergency overflow drains from the absorbers are also routed to the filtrate sump for return to the system.

- The thickener has about 1 m of sidewall above the overflow weir instead of the usual 15 to 20 cm; and the thickener hold tank is as high as the thickener. This increased sidewall height in the thickener and hold tank permits temporary storage of liquor from the other process vessels in the system, if required during maintenance; and also prevents any short-term spills of liquor due to water imbalances resulting from extreme process upsets. The total capacity provided by this additional height equals the total capacity of all the other process vessels in the system.

- The thickener hold tank is equipped with an emergency overflow tank which starts filling when extreme overflow levels are reached in the thickener and hold tank. Immediate corrective action is necessary when the overflow tank begins to fill.

ORIGINAL AND REVISED OPERATING CONDITIONS

The original design and operating conditions for the system, as shown in Table 4, specified pHs of:

- 6.0 for the absorber,

- 11.5 for the secondary reactor,

- 12 for the thickener overflow, and

- 7 to 10.5 for the top tray feed.

Unfortunately, these conditions were ideal for dissolving aluminum, a carbide lime impurity, in the reactors and thickener and then precipitating the alumina-silica complex on the absorber trays. As a result the operating pH conditions were changed from those originally specified to those shown in Table 5 and summarized below:

TABLE 4. INITIAL SET POINTS FOR SCRUBBER OPERATION

	Normal Setpoint	Allowable Range
Absorbers		
Bleed pH	6.0	6.0-6.4
Recycle to Trays (liters/min.)	2460	2300-3000
Top Tray Feed pH	N/A	7-10.5
Reheaters (T, C°)	28	20-30
Number of Modules in Service	(1) <120 MW (2) >120 MW	
Reactors		
Lime Slurry Feed Solids	25%	20%-30%
Secondary Reactor pH	11.5	11.0-11.8
Thickener Overflow pH	12.0	11-12
Number of Operating Trains (assumes 1750 ppm to 2500 ppm inlet SO_2)	(1) <150 MW (2) >150 MW	
Filters		
Drum Speed (setting)	2-1/2	0.7 rpm to give 1 cm thick cake
Wash Water (liters/min.)	230-260	not less than 230 liters/min.
Other	No wash nets Not more than 1-3 minor cloth holes Two filters operating min. of 16 hrs each day	
Soda Ash		
Makeup Rate	To maintain Na^+ Active = 0.40-0.55 M	
Makeup Solution	Use hold tank liquor	
Makeup Flow	190 liters/min.	
Makeup Feed Location	Thickener	

Source: Arthur D. Little, Inc., memorandum to LG&E dated 11/29/79.

TABLE 5. DESIGN AND REVISED OPERATING CONDITIONS FOR LG&E D/A SYSTEM

Component	Design Operating Conditions	Revised Operating Conditions
Absorber		
Flow Rate to Absorber (liters/min.)	1060 to 7000	
Spent Liquor pH	6.0	6.3
Top Tray Feed pH	7.5 to 10.5	
Primary Reactors		
Flow Rate (liters/min.)	980 to 6400	
Retention Time	3 to 15 min.	
pH	5 to 11	8.0
Secondary Reactors		
Flow Rate (liters/min.)	980 to 6400	
Retention Time	30 to 60 min.	
pH	11 to 12	9.0
Thickener		
Flow Rate (liters/min.)	2300 to 13600	
Retention Time	8.5 to 5.1 hrs.	
pH	12	<10.5
Filters		
Drum Speed	0.8 to 1.2 rpm	0.7 rpm
Wash Rate (liters/min.)	150 to 300	230 to 260
Cake Thickness (cm)	0.6 to 1.0	

- o 6.3 for the absorber spent liquor,
- o 8.0 for the primary reactor,
- o 9.0 for the secondary reactor, and
- o <10.5 for the thickener overflow.

The absorber liquor pH was raised to compensate for the lower pH of the thickener overflow fed to the top tray (and to maintain SO_2 removal efficiency). This was accomplished by incresing the "feed forward" flow of liquor from the thickener to the absorber.

It should be noted that the precipitation of the alumina-silica complex is a problem unique to LG&E Cane Run No. 6 rather than to all D/A systems, and was due to the use of carbide lime, an impure by-product of acetylene manufacture.

Formation of scale deposits on the absorber trays had been evident from start-up. The trays were recleaned whenever an outage of the boiler or scrubbing system permitted. When scrubber operation resumed, so did scale formation. Eventually enough of the tray holes would become blocked and pressure drop across the scrubber would become high enough to require a shutdown for tray cleaning.

Analyses of the tray deposits showed that the scale was mainly a complex of alumina and silica. Subsequent testing demonstrated that alumina in the carbide lime dissolved at a pH of 11.4 and precipitated when the pH dropped below 10.7.

SYSTEM PERFORMANCE DURING ACCEPTANCE TEST AND DEMONSTRATION PERIOD

The acceptance test was a comprehensive 12-day test (10 days originally scheduled) conducted during the period of July 17 to July 28, 1980, and was aimed at evaluating the performance of the D/A system for the purpose of acceptance of the system by LG&E and U.S. EPA from the system designer. During the demonstration period, which took place during the 13 months of May 1980 to June 1981, a long-term performance evaluation of the system was conducted to fully demonstrate the use of D/A FGD technology for full-scale utility applications, and prove its capabilities.

The D/A system's ability to comply with the performance guarantees was closely monitored during the acceptance test and throughout the demonstration period. SO_2 emissions were continuously measured. Filter cake samples were analyzed periodically to determine lime and soda ash consumption and the insoluble solids content of the filter cake. Power consumption was directly monitored.

Acceptance Test Results

During the acceptance test the performance of the D/A system was assessed in terms of the process guarantees. Table 6 summarizes those guarantees and the corresponding results of the acceptance test. As shown in Table 6, the FGD system did meet all the acceptance test process guarantees, with the exception of the filter cake properties. Table 7 summarizes system performance on a daily basis.

The acceptance test began at 0000 hours on July 17, 1980. For each test day, the boiler load was raised to its maximum achievable level (240-250 MW) sometime between 0600 and 0800 hours. The load was limited to this level because the FGD system booster fan could handle no more gas volume. Since the boiler was operated with an absorber inlet oxygen concentration of 7.7% (instead of the design of 5.9%), a larger volume of gas was produced at the corresponding generation rate. The system was operating at maximum design gas flowrate (about 28,000 m^3/min), yet only at 80% of design generating capacity. High load was held daily for 10 hours (an acceptance test prerequisite) and then reduced in steps to a low level of about 80 MW for 6-7 hours, generally between 0000 and 0700 hours, in line with electrical energy demand.

The test proceeded smoothly until July 23 when wet coal, resulting from heavy rains during the previous night, blocked the coal feeders to the mills. The boiler load was restricted to 150-160 MW for 2 days. Process stream analysis and stack sampling continued because the FGD system was performing well, but the test program was extended by 2 days to record 10 days of operation at full capacity. By July 25 the wet coal problems subsided, and the program continued to completion at 2400 hours on July 28.

Demonstration Period Test Results

Table 8 defines monthly performance parameters for FGD system utilization, availability, reliability, and operability. The system performance in terms of these parameters is shown in Table 9. As Table 9 shows, the FGD system met all performance guarantees except those for filter cake insoluble solids and sodium consumption. These two guarantees were not met primarily because of the filter operating problems.

The power consumption guarantee states that at gross peak boiler load the FGD system will consume a maximum of 1.2% of the power generated by the unit. Many of the power consumption figures shown in Table 9 exceed 1.2%. This is expected, since power consumption is not a linear function of boiler load, and the values shown represent periods of operation well below peak load. (Table 6 indicated that FGD power consumption was only 1.06% at 240 MW, 80% of gross peak load.)

TABLE 6. PERFORMANCE GUARANTEES AND ACCEPTANCE TEST RESULTS

Guarantee	Test Results
SO_2 Emissions Limit 200 ppm (dry basis) without additional air dilution	141 ppm (dry basis) without additional air dilution (93.9% SO_2 removal)
Calcium Consumption 1.05 moles available CaO per mole SO_2 removed	1.04 moles available CaO per mole SO_2 removed
Soda Ash Consumption 0.045 moles Na_2CO_3 per mole SO_2 removed	0.042 moles Na_2CO_3 per mole SO_2 removed
Net Particulate Addition No net particulate addition by FGD system	Net particulate removal efficiency averaged 88%
Power Consumption System will consume (excluding reheat) not more than 1.2% of power generated at gross peak boiler capacity (300 MW)	System consumed 1.06% of power generated at highest actual boiler operating load (240 MW)
Filter Cake Properties Filter cake will contain a minimum of 55-wt % insoluble solids	Filter cake averaged 52.2-wt % insoluble solids
System Availability System will have an availability of at least 90% for 1 year[a]	Not applicable to acceptance test

[a]Later modified to require an operability of 90% or greater, coupled with a daily average SO_2 removal efficiency of 90% or greater.

TABLE 7. ACCEPTANCE TEST RESULTS[a]

Acceptance Test Day	Stack SO_2 Emissions (ppm)	SO_2 Removal (%)	Particulate Removal (%)	CaO Consumption[c]	Na_2CO_3 Consumption[d]	Power Consumption[e]	Insoluble Solids (wt %)	System Availability (%)
July 17	130	94.7	85.6	1.082	0.037	1.05	52.65	100.0
July 18	129	95.0	81.7	1.054	0.038	1.08	52.20	100.0
July 19	130	94.6	89.0	1.093	0.045	1.03	52.60	100.0
July 20	152	93.4	88.2	1.096	0.031	1.07	53.72	100.0
July 21	157	93.1	90.0	1.082	0.029	1.08	53.92	100.0
July 22	140	94.5	91.8	1.044	0.070	0.99	51.90	100.0
July 23	132	93.6	85.6	1.029	0.038	1.04	50.70	100.0
July 24	124	94.1	86.0	1.018	0.032	1.07	51.40	99.6
July 25	146	93.8	88.9	0.983	0.068	1.07	51.00	100.0
July 26	171	92.7	86.9	0.995	0.028	1.08	52.43	100.0
July 27	156	93.1	88.0	1.011	0.047	1.08	52.82	100.0
July 28	130	94.0	—	0.990	0.041	1.06	50.58	100.0
Average	141	93.9	87.5	1.040	0.042	1.06	52.16	100.0

[a] Over the 12-day test period, the liquid-to-gas (L/G) ratio ranged from 1.0 to 1.3 liters/m^3; the active sodium ranged from 0.40 to 0.54 molar.

[b] Values shown are 24-hour average SO_2 results.

[c] Values shown in available CaO consumption per mole SO_2 absorbed (i.e., 1.139 total moles CaO/mole SO_2 in filter cake x 0.95 moles available CaO/total moles CaO in carbide lime feed = 1.082 moles available CaO consumed per mole SO_2 removed).

[d] Moles Na_2CO_3 consumed per mole SO_2 removed.

[e] FGD power consumption during peak boiler load.

TABLE 8. DEFINITIONS OF PERFORMANCE PARAMETERS[a]

Period Hours (A)	Hours in the period under consideration
Boiler Hours (B)	Hours during the period that the boiler was actually operated
FGD Hours (C)	Hours during the period that the FGD system (one or both modules) was actually operated
Available Hours (D)	Hours during the period that the FGD system was capable of being operated
Called Upon Hours (E)	Hours during the period that the FGD system was demanded (or wanted) for operation
Boiler Utilization ([B/A] x 100)	Hours that the boiler was actually operated divided by the period hours multiplied by 100
FGD Utilization ([C/A] x 100)	Hours that the FGD system was actually operated divided by the period hours multiplied by 100
FGD Availability ([D/A] x 100)	The hours that the FGD system was available for operation divided by the period hours multiplied by 100
FGD Reliability ([C/E] x 100)	The hours that the FGD system was actually operated divided by the called upon hours multiplied by 100
FGD Operability ([C/B] x 100)	The hours that the FGD system was actually operated divided by the boiler hours multiplied by 100

[a] Source: Edison Electric Institute.

TABLE 9. SUMMARY OF SYSTEM PERFORMANCE IN TERMS OF PERFORMANCE GUARANTEES[a]

	SO_2 Emissions[b] (ppm)	SO_2 Removal (%)	CaO Consumption[c]	Na_2CO_3 Consumption[d]	Power Consumption (%)	Insoluble Solids (wt %)	System Availability (%)
May 1980	130.0	93.2	1.23	0.079	1.10	55.0	99.9
June 1980	115.5	95.0	1.20	0.085	1.15	47.8	99.9
July 1980	143.0	94.4	1.04	0.042	1.05	52.2	99.9
August 1980	131.0	93.6	---e	---e	---	---e	99.3
September 1980	157.5	92.6	1.15	0.058	1.26	49.7	94.0
October 1980	150.0	89.8	1.04	0.076	1.43	46.8	99.9
November 1980	312.0	82.6	0.89	0.132	1.57	36.9	85.1
December 1980	---f	---f	---f	---f	---f	---f	60.0
Janmuary 1981	205.0	91.2	1.04	0.066	1.85	49.1	84.0
February 1981	129.8	93.7	1.06	0.065	1.3	50.3	99.6
March 1981	155.7	92.0	1.04	0.069	1.7	50.5	99.8
April 1981	143.8	93.1	0.93	0.045	1.18	52.1	98.6
May 1981	204.9	88.2	0.95	0.068	---	46.1	96.8
Average	164.9	91.6	1.05	0.071	1.36	48.8	93.2g
Process Guarantees	200.0	---	1.05	0.045	1.20	55.0	90.0

[a] Particulate matter removal efficiency was determined during acceptance test only (88%).
[b] All values shown (except May 1980) are 24-hour average stack SO_2 emissions, corrected for moisture and reheater. May 1980 results represent the average of the 2 absorbers' outlet emissions.
[c] Total moles available CaO consumed per mole SO_2 removed.
[d] Moles Na_2CO_3 consumed per mole SO_2 removed.
[e] No filter cake samples were analyzed in August 1980.
[f] FGD system operated only 14.4 hours in December. No system characterization tests were performed.
[g] Average availability based on hours in total period (see Table 8).

During November 1980, the system met the lime consumption guarantee, but did not meet the guarantees for SO_2 removal, availability, soda ash consumption, power consumption, and filter cake composition. The poor performance was largely attributed to the presence of significant amounts of suspended solids in the thickener overflow. Extended periods of operation at low boiler load and high oxidation rates caused the ratio of sulfate:sulfite in the scrubbing liquor to exceed 4:1. At this level of sulfate, the settling characteristics of FGD solids deteriorated, and a significant volume of FGD solids were carried over the thickener weirs and back into what should have been the clear liquor section of the process. Solids carryover from the thickener overflow increased the rate of absorber plugging and reduced the SO_2 removal capability. Lime consumption was extremely low, but this was at the expense of high consumption of soda ash, a more costly material. The high soda ash consumption resulted from reduced washing of the filter cake.

In December 1980, the FGD system operated for only 14 hours of the 240 hours that the No. 6 boiler was in operation, and none of the performance guarantees were met. The boiler was shut down primarily for scheduled maintenance, and during this outage the FGD system was considered available. However, after the boiler had come back on-line, a thickener rake problem rendered the FGD system inoperable for about 300 hours. The rake in the thickener had separated from the drive shaft. The FGD system and the thickener were briefly operated (14 hours) to keep the thickener slurry from freezing and were then shut down so the rake could be repaired. No system characterization tests were performed that month.

The four monthly dependability performance parameters defined in Table 8 are shown in Table 10 for the LG&E D/A system during the demonstration period.

CAPITAL AND OPERATING COSTS

Installation of the D/A system at Cane Run Unit 6 required capital investment in three subsystems: FGD, lime slurry feed, and waste processing and disposal. Each subsystem involved independent design and installation efforts, and the associated costs are shown in Table 11. The costs shown were reported on an as-incurred basis and thus do not represent a constant dollar value of the capital investment. The actual cost of $19,535,200 (actual costs incurred during 1977-1979 and reported in 1978 dollars) compares with the original estimate for capital costs of $17,379,000 estimated in 1978 (1977 dollars). The material costs for the FGD system and field supervision were slightly above the original estimate, while erection labor overhead and the sludge disposal system were nearly double. (However, the original estimate for sludge disposal was based on a different system that was not used.)

TABLE 10. SUMMARY OF MONTHLY PERFORMANCE PARAMETERS FOR FGD SYSTEM

	Utilization	Availability	Reliability	Operability
May 1980	99.5	99.9	99.9	99.9
June 1980	88.9	99.9	99.9	99.7
July 1980	83.1	99.9	99.9	99.8
August 1980	84.5	99.3	94.6	94.6
September 1980	86.9	94.0	98.8	98.6
October 1980	99.9	99.9	99.9	99.9
November 1980	55.1	85.1	96.8	96.8
December 1980	1.9	60.0	6.0	6.0
January 1981	80.2	84.0	83.3	83.3
February 1981	71.6	99.6	99.5	99.5
March 1981	71.9	99.8	99.7	99.7
April 1981	82.6	98.6	98.3	98.3
May 1981	96.8	96.8	96.8	96.8
Demonstration Period[a]	77.4	93.2	94.1	94.0

[a]Demonstration period averages were calculated for the entire period.

TABLE 11. ACTUAL CAPITAL COST BREAKDOWN BY SUBSYSTEM FOR CANE RUN UNIT 6 (1977-79 COSTS IN 1978 $)

	Erection Costs					Engineering Costs	
Subsystem	Material Costs[a]	Direct Labor	Field Supervision and Engineering[b]	Construction Overhead	System Supplier	LG&E Consultant	Total
FGD	10,183,200	2,386,400	277,900	1,373,000	833,500	670,000	15,724,000
Lime Slurry[c]	460,600	118,300	9,100	112,000	--	--[d]	700,000
Waste Disposal	2,084,200	387,400	32,500	398,700	208,400	--[d]	3,111,200
Total	12,728,000	2,892,100	319,500	1,883,700	1,041,900	670,000	19,535,200

a Includes spare parts at 1% of total costs.
b Owner's engineering is included in field supervision and engineering.
c Reported as a portion of the overall lime slurry system for Units 4, 5, and 6.
d Included in field supervision and engineering.

Actual annual operating costs for the D/A system were much higher than estimated, both in total annual cost and mills per kWh. The total annual operating cost estimated in 1979 was $5.1 million (with reheat), compared with the $7.8 million actually incurred (Table 12). The major reasons for this difference were higher labor, maintenance, and soda ash requirements, as well as soda ash and fuel oil unit costs that were about twice those estimated. In addition, the average boiler load was less than originally predicted. Since some operating requirements (such as labor and power) remain constant despite boiler load, operation at lower boiler load increases the cost per kWh for these components. Reduction of maintenance costs, together with improved sodium recovery and an increased unit operating load factor of 60%, could lower the operating cost from 6.9 to 5.0 mills/kWh. This figure would be comparable to that reported for the D/A unit at Southern Indiana Gas and Electric Company (5.0 mills/kWh), but much less than the 13.9 mills/kWh reported for the Central Illinois Public Service D/A installation. Overall, the capital and operating costs for the system at LG&E Cane Run No. 6 compare very favorably with the costs for other wet scrubbing processes in the U.S.

TABLE 12. ACTUAL AVERAGE ANNUAL OPERATING COSTS (WITH REHEAT) FOR CANE RUN UNIT 6 D/A FGD SYSTEM[a] (May 1, 1980 - April 30, 1981)

	Quantity[b]	Unit Cost ($)	Total Costs ($)
Direct Costs			
Carbide Lime	44,647 ton	15.32/ton	683,992
Soda Ash	5,438 ton	134.53/ton	731,574
Fuel Oil	1,204,279 gal.	1.02/gal.	1,228,365
Electricity	12,925,000 kWh	0.012/kWh	155,100
Water	79,170,000 gal.	0.05/1,000 gal.	3,959
Sludge (Trucking & Landfill)	117,890 ton	2.50/wet ton	294,725
Maintenance Materials			506,280
Labor			
Operation	83,670 hr	8.45/hr	707,012
Maintenance	71,820 hr	8.55/hr	614,061
Analysis	2,080 hr	12.00/hr	24,960
Supervision	2,080 hr	various	52,000
Total Direct Costs			5,002,028
Indirect Costs			
Overhead (Payroll & Plant)	52.9% of 1,398,033 (Total Labor)		739,560
Interest	6.125% of 19,535,200		1,196,531
Depreciation	4.65% of 19,535,200		908,387
Total Indirect Costs			2,844,478
Total Annual Operating Cost, $			7,846,506
Actual Operating Cost (at 44% load factor)			
Mills/kWh			6.9
¢/10^6 Btu			69.2
$/ton of S removed			492.5
Adjusted Operating Cost (projected for 60% load factor)			
Mills/kWh			5.8
¢/10^6 Btu			58.1
$/ton of S removed			411.3

[a] Basis: 300 MW (gross peak capacity) existing coal-fired plant. 9,960 Btu/kWh, 3.5% S coal, 91.6% SO_2 removal, average SO_2 removed - 279 lb/min. at 300 MW (gross peak) load, on-site solids disposal by trucking after treatment, stack gas reheat 50°F.

[b] Quantities of material are shown here in English rather than in metric units since material was purchased in the U.S., and these units are commonly used commercially.

Relevant conversion factors are: 1 gal. = 3.785 liters; 1 ton = 0.907 metric ton; 1 Btu = 1055 J; 1 lb = 0.454 kg; °F = (1.8) (°C)+32

BIBLIOGRAPHY

Chang, J.C.S. (Acurex Corporation), N. Kaplan. "SO_2 Removal by Limestone Dual Alkali." IERL-RTP-P-630. For publication in Environmental Progress, 1984.

Chang, J.C.S. (Acurex Corporation), N. Kaplan. "SO_2 Removal by Limestone Dual Alkali." Paper presented for handout at EPA/EPRI Symposium on Flue Gas Desulfurization, New Orleans, LA, November 1983.

Chang, J.C.S. (Acurex Corporation), N. Kaplan. "Pilot Evaluation of Limestone Regenerated Dual Alkali Process." Paper presented at EPA/EPRI Symposium on Flue Gas Desulfurization, New Orleans, LA, November 1983.

VanNess, R.P. (Louisville Gas & Electric Company), L.R. Woodland, E.D. Gibson (Arthur D. Little, Inc.). "Full-Scale Dual Alkali FGD Demonstration at Louisville Gas and Electric Company." EPA-600/7-83-039 (NTIS PB 83-241 364), August 1983.

Watson, D.A., T.M. Martin, J.K. Donnelly, N. Zoueshtiagh (Bechtel National, Inc.), R.P. VanNess (Louisville Gas & Electric Company), C.R. LaMantia, L.R. Woodland (Arthur D. Little, Inc.). "Dual Alkali Acceptance Test at Louisville Gas and Electric Company -- Volume I. Acceptance Test and Appendices A-C (NTIS PB 82-231 267), Volume II. Appendices D-F (NTIS PB 82-231 275), Volume III. Appendices G-J (NTIS PB 82-231 283), EPA-600/7-81-159a, b, c," September 1981.

Valencia, J.A., R.R. Lunt (Arthur D. Little, Inc.), G.J. Ramans (Thyssen-CEA Environmental Systems, Inc.). "Evaluation of the Limestone Dual Alklai Prototype System at Plant Scholz: System Design and Program Plan." EPA-600/7-81-141a (NTIS PB 81-247 421), August 1981.

Valencia, J.A., J.F. Peirson, Jr. (Arthur D. Little, Inc.), G.J. Ramans (Thyssen-CEA Environmental Systems, Inc.). "Evaluation of the Limestone Dual Alklai Prototype System at Plant Scholz: Final Report." EPA-600/7-81-141b (NTIS PB 82-110 685), August 1981.

Gibson, E.D., T.S. Hurley, J.C. Lacy (Radian Corporation). "Summary Report: Sulfur Oxides Control Technology Series: Flue Gas Desulfurization - Dual Alkali Process." EPA-625/8-80-004, October 1980.

VanNess, R.P., R.C. Somers, R.C. Weeks (Louisville Gas & Electric Company), T. Frank, G.J. Ramans (Combustion Equipment Associates), C.R. LaMantia, R.R. Lunt, J.A. Valencia (Arthur D. Little, Inc.). "Executive Summary (Report) for Full-Scale Dual-Alkali Demonstration System at Louisville Gas and Electric Co. - Final Design and System Cost." EPA-600/7-79-221a (b), (NTIS PB 80-146 707 (80-146 715)), September 1979.

VanNess, R.P., R.C. Somers (Louisville Gas & Electric Company), T.M. Frank, J.M. Lysaght (Combustion Equipment Associates), I.L. Jashnani, R.R. Lunt, and C.R. LaMantia (Arthur D. Little, Inc.). "Project Manual for Full-Scale Dual-Alkali Demonstration at Louisville Gas and Electric Co. - Preliminary Design and Cost Estimate." EPA-600/7-78-010 (NTIS PB 278722), January 1978.

Kaplan, N. "Introduction to Double Alkali Flue Gas Desurization Technology." Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, March 1976, Vol. I, EPA-600/2-76-136a (NTIS PB 255317), pp. 387-421, May 1976.

LaMantia, C.R., R.R. Lunt (Arthur D. Little, Inc.), R.E. Rush (Southern Services, Inc.), T.M. Frank (Combustion Equipment Associates, Inc.), N. Kaplan. "Operating Experience--CEA/ADL Dual Alkali Prototype System at Gulf Power/ Southern Services, Inc. Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, March 1976, Vol. I, EPA-600/2-76-136a (NTIS PB 255317), pp. 423-469, May 1976.

Kaplan, N. "An Overview of Double Alkali Processes for Flue Gas Desulfurization." Proceedings: Symposium on Flue Gas Desulfurization, Atlanta, November 1974, Vol. I, EPA-650/2-74-126a (NTIS PB 242572), pp. 445-515, December 1974.

LaMantia, C.R., R.R. Lunt, J.E. Oberholtzer, E.L. Field (Arthur D. Little, Inc.), Kaplan, N. "EPA-ADL Dual Alkali Program--Interim Results." Proceedings: Symposium on Flue Gas Desulfurization, Atlanta, November 1974, Vol. I, EPA-650/2-74-126a (NTIS PB 242572), pp. 549-665, December 1974.

Kaplan, N. "An EPA Overview of Sodium-Based Double Alkali Processes, Part II, Status of Technology and Description of Attractive Schemes." Proceedings: Flue Gas Desulfurization Symposium, 1973, EPA-650/2-73-038 (NTIS PB 230901), pp. 1019-1060, December 1973.